Separation Science and Technology
Publication details, including instructions for authors and subscription information:
http://www.informaworld.com/smpp/title~content=t713708471

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To cite this Article Goldberg, Morris and Rubin, Eliezer(1972) 'Foam Fractionation in a Stripping Column', Separation Science and Technology, 7: 1, 51 — 73

To link to this Article: DOI: 10.1080/00372367208058971

URL: http://dx.doi.org/10.1080/00372367208058971

SEPARATION SCIENCE, 7(1), pp. 51–73, 1972

Foam Fractionation in a Stripping Column

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Summary

Experiments with stripping foam fractionation columns indicate that within the range of variables studied, stripping lengths of 10 to 150 cm have negligible effect on separation. These experimental results cannot be explained adequately by the transfer unit approach.

A physical model is proposed wherein the stripping length consists of two mixing regions and a countercurrent flow region in between. Negligible solute transfer is assumed in the countercurrent flow region. This model is applied to explain the experimental data.

Foam fractionation is a separation method utilizing foam as a medium of large specific interfacial area for partial separation of components of a solution containing surface-active solutes.

Surface-active solutes adsorb to the gas–liquid interface. In a surfactant solution, a large portion of the solute can therefore be removed together with the gas–liquid interface by continuous formation and removal of an interface. Fortunately, many surfactant solutions form foams which can be easily produced and removed. Thus the same chemicals that concentrate at the interface often have the ability to form large surfaces which offer the means for affecting an efficient separation.

Several extensive reviews of the foam fractionation process and its

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applications have appeared (*1, 2*). Past attempts at modeling continuous foam fractionation included two types of columns: single stage (or simple) and stripping columns. In a simple column, feed enters the liquid pool under the foam, and exiting foam is in equilibrium with the relatively low concentration of the bottoms liquid pool. In a stripping column, feed is entered directly to the foam. The exit foam can then be in equilibrium with a higher concentration than that of the bottoms liquid, and a higher degree of separation is possible. For stripping operation, equations have been published describing mathematically infinite columns where distance between the bottoms pool and the liquid entry into the foam is large (*3–5*). For short stripping columns equations have been written describing mass transfer between countercurrent streams within the foam liquid (*6, 7*).

In reported experimental data on stripping columns, either the effect of column lengths has not been investigated fully (*5*) or the inadequacy of the theoretical equations has been ascribed to detrimental flow effects (*7*). In any case, the theory to date has not been successful in predicting the behavior of short stripping columns.

This work attempts to compare two models of stripping foam fractionation columns. One model is based on the hitherto assumed continuous solute transfer between countercurrent streams within the foam, and the other on "end effects" with hardly any solute transfer within the foam. The latter model, corresponding to new experimental results, predicts that to a large extent separation should be independent of stripping column length (or height).

THEORY

Any theory on foam fractionation columns has to consider foam structure and liquid flow within the foam. Pertinent properties of liquid foam and references are therefore indicated briefly in the following section.

Foam Properties

Reviews of the theory and properties of foams have been presented by Kitchener and Cooper (*8*) and Bikerman (*9*). Foam consists of gas bubbles distributed in a liquid medium. Wet foams contain round gas bubbles separated by fairly thick walls, with the bubbles substantially far apart and not appreciably deformed. In dry foam the bubbles are separated by thin films and have a polyhedral shape. The films in general meet at an angle of 120°, and their line of contact is known as a plateau border (PB). This is a symmetrical channel bounded by sections of cylinders of radius usually much less than the bubble diameter.

Drainage of liquid from static foams has received much attention, with many earlier works based on studies of single films. Mysels, Shinoda, and Frankel (*10*) distinguished between rigid and simple mobile films. The latter drain mainly through a process that they call marginal regeneration, where entire sections of film are pulled into an adjacent PB because of its relatively low pressure and are replaced by a thinner film. Marginal regeneration results in a net transfer of liquid into the PB's. The film eventually reaches a minimum steady-state thickness. Aqueous solutions of the common surfactants form mobile films.

Moving or dynamic foams have been studied only in recent years. Dynamic foams are characterized by plug flow regime at relatively low linear velocities and liquid content, and turbulent flow regime at higher linear velocities and liquid content (*11*). Continuous stripping columns usually have to be operated in the plug flow regime since any turbulence results in considerable longitudinal mixing which is detrimental to the multistage effect.

The work of Mysels et al. implies that in a well-drained mobile film of the type to be expected in the present work any flow of liquid takes place through the PB's of the foam and not through the bubble faces. Various models of foam flow and of liquid flow within moving and stationary foams have been published, based on various ideal configurations of the PB's, which were considered to contain the major portion of the liquid flow. Haas and Johnson (*12*) pictured cylindrical capillary tubes with elastic walls, and Leonard and Lemlich (*13*) used a more nearly correct curved triangular cross section, performing calculations numerically with the aid of a digital computer. This second model was recently further refined (*14*).

All theoretical models to date, including those presented below, are restricted to the plug flow regime.

Concentrations within the Foam

Concentrations within the foam are conveniently described using the Guggenehim concept of surface activity in an idealized model of the foam (*1*). The air–solution system is divided into three phases: *l*, the

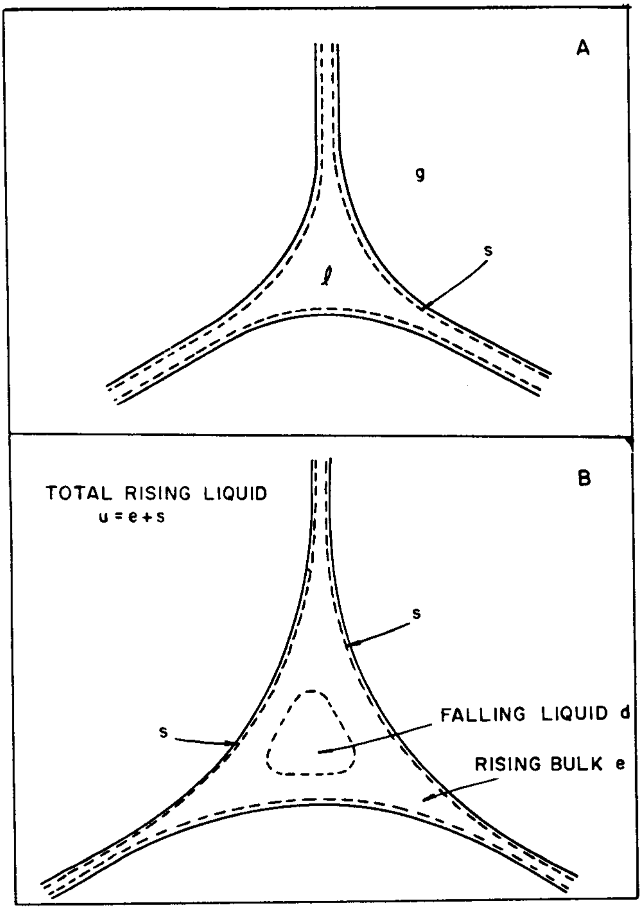

FIG. 1. Phases in plateau border and associated films: (A) uniform bulk (in simple column), (B) nonuniform bulk (in stripping column).

uniform bulk phase containing most of the liquid; *g*, containing all of the inert gas; and *s*, the interfacial phase. The *g*–*s* boundary is taken at the physical interface, and the *l*–*s* boundary must be far enough from the interface so that all gradients take place within the *s* phase.

Figure 1A is an axial view of a PB, with the phase boundaries shown. The foam liquid can also be represented by a single liquid film, as shown in Fig. 2. As indicated in the figure, all of the liquid is taken into account in the foam ratio *f*, defined as volume liquid/volume foam. Similar ratios are defined for each phase individually. For instance, f_s is the volume of liquid in phase s/volume foam. In Figure 2B, the solute concentration profile in the foam liquid is illustrated, and the average concentrations within the liquid are also indicated. For instance, y is the average solute concentration in the foam liquid.

For this model, the bulk and average concentrations can be related by

$$y = (a\Gamma/f) + x_l \tag{1}$$

The term $a\Gamma/f$ is the excess solute based on unit volume of total liquid in the foam. Equation (1) applies at any point in a foam column as long as the restrictions with respects to phase boundaries are met, i.e., the bulk concentration x_l can be assigned a single value throughout the l phase. It is also assumed that equilibrium exists between the bulk and

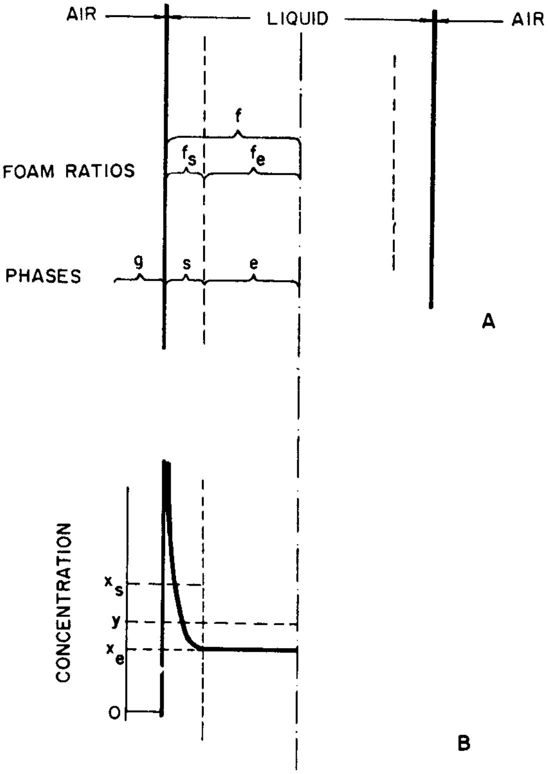

FIG. 2. Foam liquid represented by liquid film (uniform bulk): (A) schematic diagram of liquid film, (B) concentration within the film.

surface phases in the foam. The surface excess Γ appears in the Gibbs equation, which for a single solute in water usually can take the form

$$d\gamma = -R'T\Gamma d \ln (x) \tag{2}$$

Underneath a point of liquid feed (as in a stripping column), the foam liquid consists of two streams originating from different locations in the column. Figure 1B illustrates schematically an axial view of a PB under these conditions. The downward flow takes place within the PB, while the liquid in and near the foam film moves upward. The concentration in the falling liquid can in general differ from the bulk concentration in the rising liquid. Therefore, what was considered the l phase in Fig. 1A

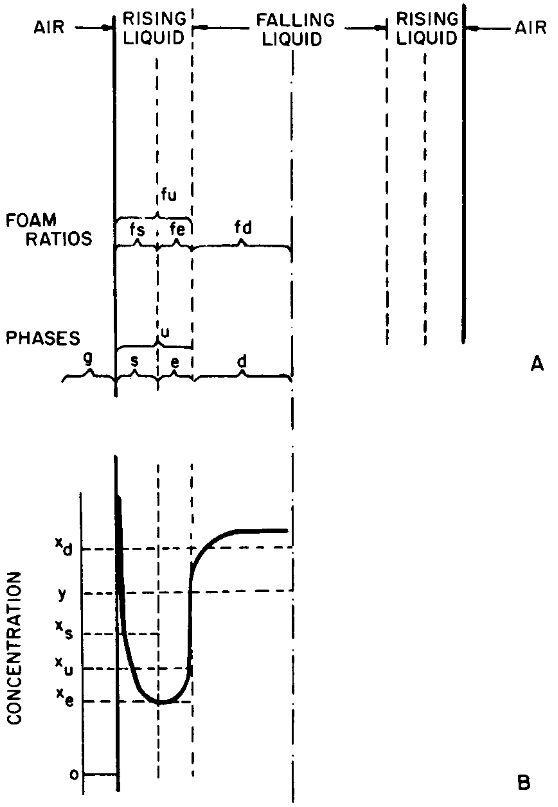

FIG. 3. Foam liquid represented by liquid film (nonuniform bulk): (A) schematic diagram of liquid film, (B) concentration within the film.

has been divided in Fig. 1B into d, the falling liquid, and e, the rising bulk liquid. Phases s and e together contain all of the rising liquid, designated u.

The foam liquid at any cross section of the column can also be represented by a single film, as shown in Fig. 3A, with concentrations and foam ratios in Fig. 3B. In view of the complicated flow pattern within the foam it is improbable that the respective solute concentrations within phases e and d will actually be uniform as shown in Figs. 3A and 3B. The uniform concentrations shown in the figure can be considered idealized mixing-cup concentrations for the liquid of each phase.

An equation analoguous to Eq. (1) can now be written where y is replaced by x_u, the average concentration in the rising liquid:

$$x_u = (a\Gamma/f_u) + x_e \tag{3}$$

Equation (3) also applies at any point in a foam column, if equilibrium can be assumed between the s and e phases. Note, however, that f_u cannot be measured experimentally and it may have to be estimated before Eq. (3) can be used.

As for concentrations and flows along the column, two idealized models can be used. According to one model it is assumed that there is appreciable solute transfer between rising and falling liquid. This model is utilized below to derive equations reported in the literature for stripping columns.

According to the second model proposed below, there is very little solute transfer between the falling and rising streams. Most of the stripping is due to "end effects" originating from mixing at the feed entry point and at the bottom of the foam. This assumption is born from considering the actual dimensions involved in Fig. 1B and the work of Mysels et al. (*10*) mentioned before. Only a small fraction of the gas–liquid interfacial area is close enough to the falling liquid to affect any appreciable concentration changes within the residence times involved in the stripping section of the foam column. Thus, unless the falling liquid is mixed with the rising liquid, solute transfer between the streams is negligible. Thus the stripping column length (i.e., the distance between feed entry point and the bottom of the foam) can be divided into three regions: A mixing region around the feed entry point, a mixing region at the liquid pool (including primarly the upper part of the liquid pool) at the bottom of the foam column, and a countercurrent region in between. Solute transfer in the countercurrent region is assumed negligible. The term "mixing region" is used here quite liberally to indicate a region of good contact between falling and rising liquid streams. The good contact around the feed entry point is due to a much wetter foam relative to the countercurrent region, and not necessarily to backmixing of the foam.

Support to this model is also found by considering the similarity between stripping foam columns and liquid–liquid spray columns. Recent studies (*15*) indicated that heat and mass transfer in spray columns is characterized by a sharp drop in the driving force (temperature or concentration jump) at the inlet of the continuous phase. These concentration and temperature jumps indicate that similar, or even more extreme, concentration jumps may exist in stripping foam columns. Therefore, only very little solute is transferred between the rising and falling liquid in the countercurrent region.

Application to Foam Columns—Models from the Literature

A simple column operating at steady state is illustrated in Fig. 4. The feed is added to the liquid pool. Such a column can be useful for foam separation in some cases and also as a convenient method of measuring the surface excess Γ. The following assumptions are made: (a) the bulk concentration x_l is constant throughout the column and is equal to x_B, (b) the bulk and surface phases are in equilibrium in all of the foam, (c) average bubble diameter is constant in the column, and (d) the foam is stable.

Equation (1) can then be applied to the foam exiting from the column, and

$$y_t = x_B + (a_t\Gamma_b/f_t) \tag{4}$$

Other relations result from applying a material balance in addition to Eq. (4). For instance,

$$x_B = x_F - (a_t\Gamma_b V_t/F) \tag{5}$$

These equations and similar ones for single-stage columns have been reported in the literature (*1*, *2*).

Figure 5 illustrates a steady-state stripping column, with the feed introduced directly into the foam. There is no longer a uniform bulk concentration x_l at all positions within the column, but if the column is sufficiently long it may be assumed that solute is transferred from phase d to phase u in the foam. For an "infinite" column concentration x_e (bulk in phase u) will then approach x_F near the feed level (f in Fig. 5). Thus it should again be possible to consider a uniform bulk

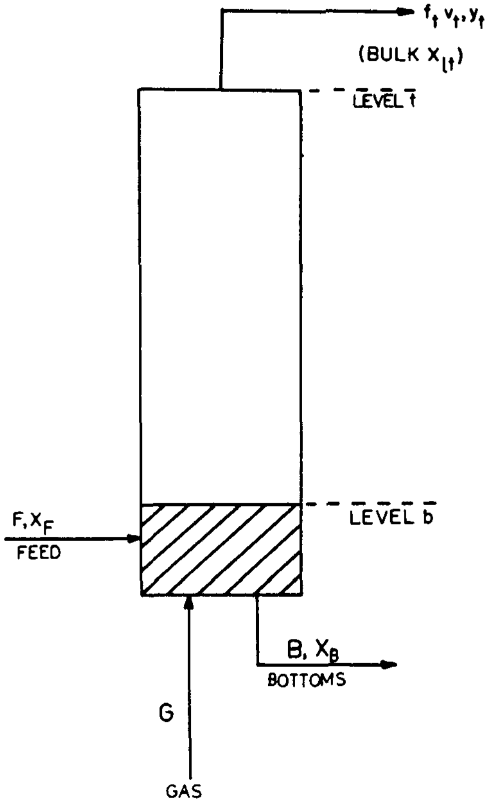

FIG. 4. Concentrations and flows in a continuous simple foam fractionation column.

concentration in the foam at the feed point. This bulk concentration $x_{lf} = x_{lt}$ will be very close to x_F. Applying Eq. (1) and a material balance leads to such equations as

$$y_t = x_F + (a_t\Gamma_f/f_t) \tag{6}$$

$$x_B = x_F - (a_t\Gamma_f V_t/B) \tag{7}$$

Equations equivalent to Eqs. (6) and (7) have been presented in the literature (*2–5*).

Relations can also be developed for "short" stripping columns, based on assumed solute transfer between the two streams in the foam liquid for the length of the column. The internal concentrations for such a

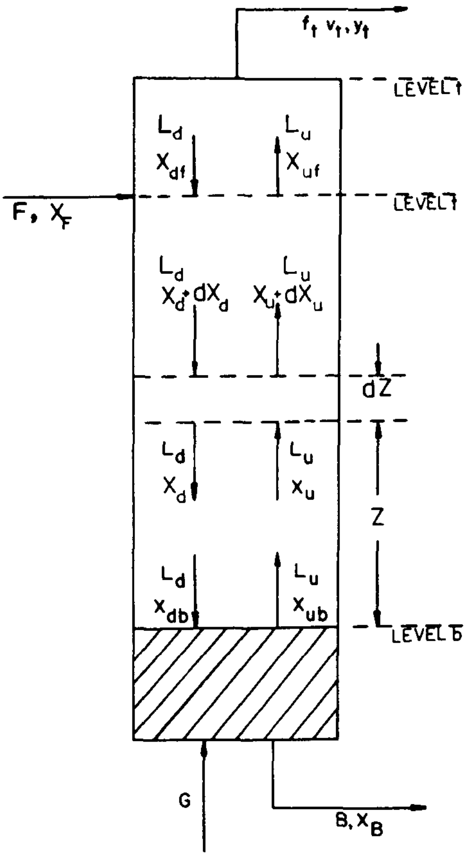

FIG. 5. Concentrations and flows in a short continuous stripping column.

column are indicated in Fig. 5. The operating line for the column results from a material balance around the bottom section of the column:

$$L_d(x_d - x_{db}) = L_u\left(x_e + \frac{a_m\Gamma}{f_m} - x_{eb} - \frac{a_m\Gamma_b}{f_m}\right) \tag{8}$$

There are two specific cases that can be treated easily. If the distribution factor $\alpha = \Gamma/x$ is constant for the range considered, Γ will change throughout the column and phase e must be negligible for Eq. (8) to be linear and for easy integration. This corresponds to assuming that solute transfers directly from phase d to the interface with an equilibrium concentration equal to Γ/α. Using the transfer unit approach and assuming that mass transfer is by molecular diffusion, stripping column height can be calculated from

$$Z = (\mathrm{NTU})(\mathrm{HTU}) \tag{9}$$

where

$$(\mathrm{HTU}) = \frac{L_d l_e}{a_m A_c D_m} \tag{10}$$

and

$$(\mathrm{NTU}) = \frac{(x_{df} - x_{db})}{(x_d - \Gamma/\alpha)_{LM}} \tag{11}$$

These equations are identical to the equations derived and tested by Haas and Johnson (*7*). They found only limited agreement between the theory and experimental results.

If, on the other hand, surface excess Γ is constant throughout the column, the interface can be considered saturated in the whole column and mass transfer involves only the internal phases *e* and *d*. In this case, Eq. (8) again takes on a linear form. Equations (9) and (10) apply, but Eq. (11) must be replaced by

$$(\mathrm{NTU}) = \frac{(x_{df} - x_{db})}{(x_d - x_e)_{LM}} \tag{12}$$

Model of Stripping Column without Solute Transfer in Countercurrent Flow Region

The assumption of negligible solute transfer within the countercurrent flow region of a stripping column leads to the conclusion that a change in stripping column height has no effect on the separation.

A stripping column as illustrated in Fig. 5 will be considered again. For the proposed model, the column can be represented by the schematic diagram of Fig. 6. Here there is no interaction between the countercurrent streams L_u and L_d. The advantage over simple column operation is a result of steady-state flow into and out of two regions of perfect mixing within the foam liquid: the liquid pool, subscript *b*; and the feed level, subscript *f*. Regions *b* and *f* are connected only through internal flows L_u and L_d.

For the extreme case of $\Gamma_t = \Gamma_F$ (i.e., in equilibrium with x_F) this

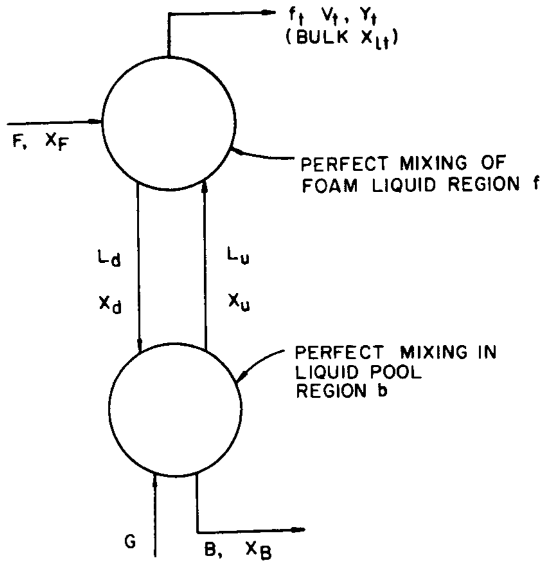

FIG. 6. Idealized liquid flow in stripping column. No solute transfer in countercurrent flow region.

model leads to equations identical to Eqs. (6) and (7). For all other cases, i.e., $\Gamma_t < \Gamma_F$, total material and solute balances for the bottom section of the column yield

$$L_d = L_u + B \tag{13}$$

$$x_d L_d = x_u L_u + x_b B \tag{14}$$

The only assumption now necessary is that the bulk liquid concentration is equal in the two streams exiting from each of the perfectly mixed regions. That is,

$$x_{eb} = x_B; \qquad x_{df} = x_{lt} \tag{15}$$

Finally, note that the internal surface area can be calculated from

$$a_m = 6(1 - f_m)/D_v \tag{16}$$

Equation (3) and Eqs. (13) to (15) can be combined to give

$$\frac{(x_{lt} - x_B)}{(6\Gamma_b/D_v)} = \frac{L_u(1 - f_m)}{L_d f_u} \tag{17}$$

If a relation between Γ and x is known, the value of x_{lt} can be calculated from a measured value of y_t by use of Eq. (1):

$$y_t = (a_t\Gamma_t/f_t) + x_{lt} \tag{18}$$

The left-hand side of Eq. (17) is a function of quantities that can thus be measured directly or calculated. The term on the right-hand side is a function of the internal flows in the countercurrent region of the column.

EXPERIMENTAL

The experimental system was designed to provide various configurations of continuous foam separation columns. The most important part of the equipment was the foam column, which was designed for versatility so that column height could be changed conveniently between wide limits, and the same equipment could be used with minimum changes for simple or stripping operation.

A schematic diagram of the experimental system, operating as a stripping column, is presented in Fig. 7. Feed was pumped from a feed stock jar to a header from which it entered the column after passing through one of two calibrated rotameters. Bottoms liquid exited through a calibrated rotameter. Foam was collapsed in a mechanical foam breaker, and the liquid then exited from the system. Filtered saturated air from the laboratory supply entered the column through a platinum disk in which holes had been drilled. The bubbles which formed collected into a foam that traveled up the column.

The basic parts of the column were interchangable sections of nominal 3-in. Pyrex pipe. Adjacent sections were attached by use of chrome-plated brass rings, described elsewhere (*11*). A stainless steel plate formed the bottom of the column. Tubes welded to the plate provided gas inlet and liquid outlet and (where needed) inlet. Liquid feed into the bottoms pool was above the level of the plate in order to prevent short-circuiting from the feed entrance to the bottoms outlet. Liquid feed into the foam for stripping operation was through a stainless steel "spider," similar to the ones discussed by Haas and Johnson (*7*). The spiders used had three or four inlet tubes.

Photographs of the foam at the column wall were taken with a Pentax Sla single-lens reflex camera. Positive prints were prepared at an enlargement of about 10× (length) natural size, and average bubble size was calculated from measurements of the enlarged pictures.

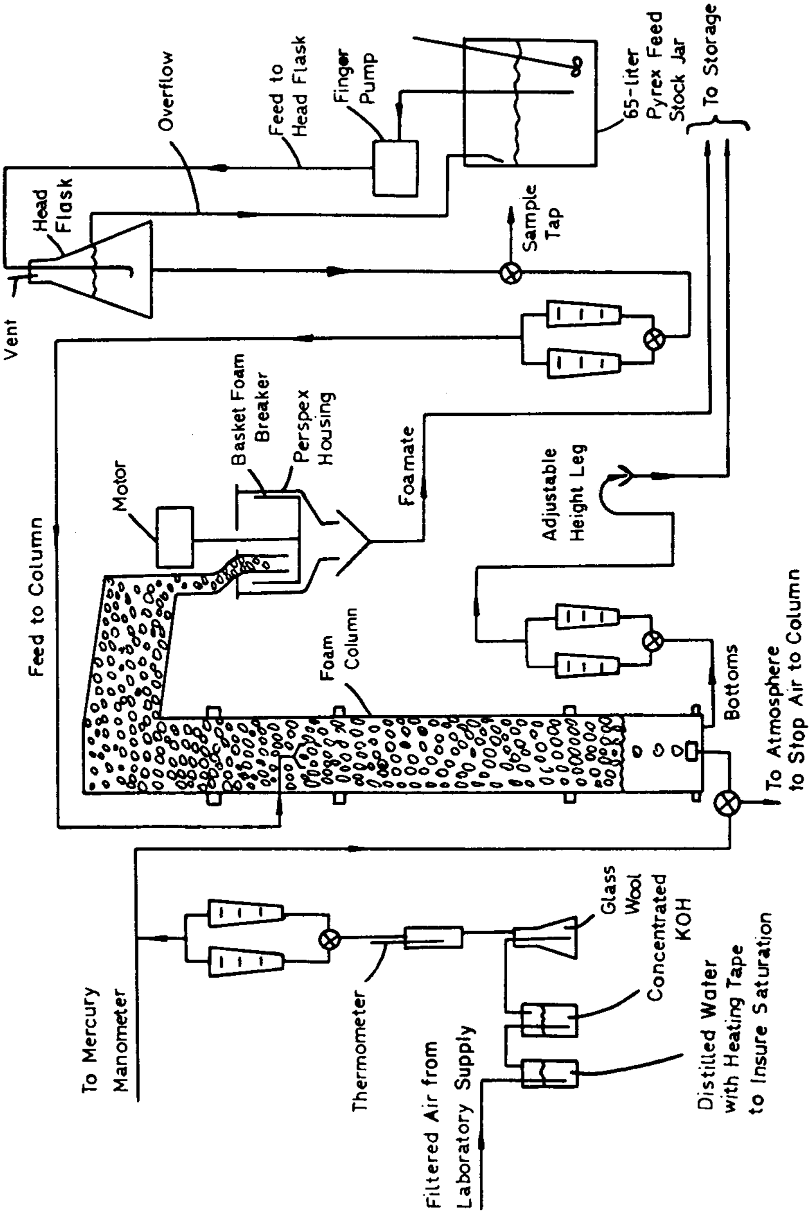

FIG. 7. Schematic diagram of experimental system.

Chemical analyses were performed with the aid of a Beckman DB spectrophotometer. Direct plots of absorbance were obtained on a Sargent Model SRL Recorder connected to the spectrophotometer.

A Fisher Scientific Co. semiautomatic Tensiomat Model 21 was used for surface tension measurements. All foam column experiments were performed with solutions in distilled water. Surface tension of the distilled water at 23°C was 71.6–71.8 dyne/cm for repeated measurements.

The surfactant used for all foam separation experiments was RWA 240, obtained from the Roberts Chemical Co. This is a 35% solution of monobutyl diphenyl monosodium sulfonate. Surface tension of RWA 240 solutions in distilled water showed no minimum in the neighborhood of the critical micelle concentration (CMC) when surface tension was plotted against concentration. This can be taken as one indication of a high degree of purity. The value of the CMC was about 6.5×10^{-3} *M*. For all foam separation experiments, the solution contained 100 mg/liter of NaCl (A.R.). Preliminary experiments showed that this helped stabilize the foam.

Surfactant was conserved by recombining exit liquid streams from the column for reuse in later runs. Chemical analysis showed that there was no detectable aging effect. However, as a safety measure, solutions were not used if they were more than about 3 weeks old.

Analysis of RWA 240 was based on spectrophotometric measurements at a wavelength of about 250 mμ where there was a pronounced peak. Samples taken from the column were diluted volumetrically 20 to 40 times. A calibration curve with known samples gave a straight line through the origin to a concentration of 2×10^{-4} *M* when absorbance was plotted against concentration.

RESULTS AND DISCUSSION

Simple Columns

Surface excess Γ was calculated from Eqs. (4) and (5), based on experimental values of flow rates, concentrations, bubble size, and exit foam ratio. The experimental values indicated no consistent difference depending on whether Eq. (4) or Eq. (5) was used. There was, however, a clear effect of concentration, with Γ increasing as x_B increases. This effect is reasonable in that the range of concentration in the present study, 1.6×10^{-3} to 5.6×10^{-3} *M*, was somewhat below the CMC.

For convenience, it was assumed that the concentration range was small enough for the data to be adequately represented by a straight line. In any case, the spread of the data justified a curve no more than a straight line. The equation of the least-squares line for the experimental data is

$$\Gamma = (0.34 \times 10^{-4}x) + (1.73 \times 10^{-7}) \tag{19}$$

Equation (19) is valid only for the concentration range encountered in the present work.

Stripping Columns

Method of Analysis of Results. The experimental foam column was operated as a stripping column for a range of height, concentration, and flow conditions. The main question to be answered by the experimental data was what is the importance of solute transfer within the foam.

One way of analyzing the results could be to calculate the experimental value of the number of transfer units as a function of stripping column height. If solute transfer within the foam is important, NTU would be expected to increase with stripping column height for a given set of column conditions. If solute transfer is unimportant, on the other hand, NTU would be independent of stripping column height for a given set of column conditions. However, in the present work Γ is given by Eq. (19), and therefore neither Eqs. (11) nor (12) can be used.

An empirical measure of separation which is equivalent to the calculation of NTU was therefore chosen. According to the theory presented above, the lowest limit of separation for a stripping column operating at steady state will be that of a simple column, where the exit foam is in equilibrium with the liquid pool. In stripping operation, separation can be greater than this value. The maximum separation will approach that of a mathematically infinite column, with the exiting foam in equilibrium with the feed solution.

For given column conditions, the separation can be calculated for simple and infinite stripping columns operating at the same flow rates and bubble size as the experimental column and with the same feed composition. A convenient measure of separation can then be the difference between maximum and minimum concentrations exiting from the column, $(y_t - x_B)$. Comparing the experimental separation achieved to the calculated separations for a simple column and for an infinite column corresponds in the present case to a calculation of NTU. For example, there are two extreme cases: (a) when NTU = 1, the column operates as a simple column, and the experimental separation $(y_t - x_B)$ is equal to that calculated for a simple column; and (b) when NTU → ∞, the column approaches infinite stripping operation, and the experimental separation approaches that of an infinite column. By calculation $(y_t - x_B)_1$ and $(y_t - x_B)_\infty$ and comparing $(y_t - x_B)_{\text{exp}}$ to these values, it will thus be possible to judge what type of operation has been achieved.

Column operation can then be analyzed by studying the effect of stripping height on experimentally obtained separation. There are three possible cases (or combinations of these), depending on which mechanism is most nearly correct: (a) For all stripping heights studied, separation is what would be calculated from simple column results. In this case, it would be concluded that feeding directly into the foam offered no advantage; (b) Experimental separation is close to $(y_t - x_B)_1$ for short columns, and increases to approach $(y_t - x_B)_\infty$ as stripping column height is increased. In this case, solute transfer within the foam can be considered important; (c) Experimental $(y_t - x_B)$ is between the values calculated for simple and infinite stripping columns at some flow conditions, and for these cases it is substantially independent of stripping column height. It would then be necessary to assume that solute transfer in the foam is not important.

Results. The necessary calculations were carried out for all stripping column runs. All flows, bubble size, exit foam ratio, and feed concentration were taken equal to the experimental values for the given run. The appropriate value of Γ was known from Eq. (19). For a simple column,

$$x_B = \frac{x_F - (1.73 \times 10^{-7})(6G/FD_v)}{1 + (0.34 \times 10^{-4})(6G/FD_v)} \tag{20}$$

The separation for a simple column was then

$$(y_t - x_B)_1 = \frac{6(1 - f_t)\Gamma_b}{f_t D_v} \tag{21}$$

with Γ_b calculated from Eq. (19).

For an infinite stripping column Γ_f could be calculated directly from Eq. (19), since x_F was known. The separation was then

$$(y_t - x_B)_\infty = \frac{6(1 - f_t)}{f_t D_v}\left(\frac{F}{B}\right) \tag{22}$$

For the experimental system, column height could not be changed conveniently during a given day's run. In order to test which of the postulated mechanisms is correct, it was necessary to duplicate conditions on successive days at various column heights. However, it was not always possible to duplicate experimental conditions exactly from one day to the next. In spite of this, it has been possible to choose several groups of experiments where the match is fairly close, so that the effect of stripping column height can be examined. Typical pertinent results are given in Table 1. The detailed results are on file (*16*).

Most of the data sets presented in Table 1 indicate that the mechanism involving poor solute transfer is most probably correct. The experimental separation is between that for simple and infinite stripping columns and is not influenced by stripping column height. Stripping column height for these experiments was between about 10 and 152 cm. Note that at heights of about 10 cm or less, there may have been a small effect of column height (see, for instance, data sets 5 and 7).

Correlation of the Results

The results were felt to be most compatible with the newly proposed model of negligible solute transfer inside the stripping section. Accordingly, the term $(x_{lt} - x_B)/(6\Gamma_b/D_v)$ was calculated for each of the stripping column results. The value of x_{lt} could be calculated from Eqs. (18) and (19), giving

$$x_{lt} = \frac{y_t - (1.73 \times 10^{-7}a_t)/f_t}{1.0 + (0.34 \times 10^{-4}a_t)/f_t} \tag{23}$$

From Eq. (17), $(x_{lt} - x_B)/(6\Gamma_b/D_v)$ should be equal to $L_u(1 - f_m)/f_uL_d$ and independent of stripping column height if solute transfer within the foam is negligible. The latter term is made up of properties of the foam inside the column. These properties, though not measurable directly, should correlate with measurable foam properties and external flows. But by Eq. (19) it should also be possible to correlate the first term, $(x_{lt} - x_B)/(6\Gamma_b/D_v)$, with external flows and readily measurable quantities. Accordingly, a least-squares regression computer program written in Algol for Technion's Elliott 503 computer was used to search for an empirical correlation between $(x_{lt} - x_b)/(6\Gamma_b/D_v)$ and foam properties, of the form:

$$\frac{(x_{lt} - x_B)}{(6\Gamma_b/D_v)} = KD_v^{p_1}G^{p_2}F^{p_3}B^{p_4} \tag{24}$$

TABLE 1

Sets of Stripping Columns at Close to Identical Conditions—Effect of Stripping Height

Set	B (cc/min)	F (cc/min)	G (l/min)	D_v (cm)	f_t (%)	x_F ($M \times 10^{-3}$)	Stripping height (cm)	$(y_t - x_B)M \times 10^{-3}$ Simple	$(y_t - x_B)M \times 10^{-3}$ Experimental stripping	$(y_t - x_B)M \times 10^{-3}$ Infinite stripping
1	24	61	2.26	0.173	1.61	5.18	25	0.71	1.71	1.87
	21	60	2.25	0.173	1.70	5.20	9	0.67	1.74	2.00
2	14	37	1.76	0.182	1.29	5.79	73	0.89	2.05	2.46
	14	42	1.77	0.178	1.56	5.95	152	0.76	1.96	2.39
	15.5	43.5	1.76	0.163	1.57	5.85	35	0.81	1.89	2.40
3	18.5	68.5	2.12	0.161	2.30	3.90	14	0.46	1.20	1.79
	16	66	2.12	0.168	2.30	3.61	35	0.43	1.08	1.85
4	20	59	2.23	0.182	1.72	3.60	35	0.53	1.13	1.69
	20	53	1.90	0.181	1.71	3.65	144	0.54	1.27	1.49
5	76	137	3.05	0.183	1.96	1.76	50	0.37	0.65	0.69
	76	138	3.10	0.176	1.96	1.76	32	0.39	0.63	0.72
	65	139	3.25	0.178	2.23	1.75	12	0.33	0.54	0.73
6	10	128	2.65	0.143	4.26	4.80	35	0.31	1.40	4.05
	12	129	2.65	0.138	4.22	4.69	73	0.32	1.39	3.51
7	10	66	2.22	0.160	2.46	3.91	4	0.43	0.96	2.99
	11	71	2.20	0.144	2.66	3.87	7	0.44	1.01	2.99
	10.5	69.5	2.19	0.150	2.62	3.88	10	0.43	1.10	2.99
8	15	140	2.65	0.138	4.52	4.80	35	0.30	1.25	2.89
	18	133	2.65	0.133	4.15	4.69	73	0.33	1.31	2.54
9	30	153	2.45	0.134	4.79	5.85	35	0.32	1.42	1.69
	25	123	2.24	0.140	4.19	5.92	152	0.35	1.52	1.79
	29	127	2.24	0.134	4.19	5.95	152	0.37	1.48	1.67

The effect of stripping height was tested by dividing the data into five arbitrary groups, depending of the stripping column height, for the following ranges, in cm: 1–2, 4–7, 9–14, 25–35, and 50–152. For the three sets with column height greater than 9 cm, the exponents of D_v and B have similar and overlapping confidence ranges. The data for the shorter columns, on the other hand, were inconsistent. All the data for stripping heights longer than 9 cm were accordingly grouped together, and a regression of the form of Eq. (24) was tested for the combined data. Only the exponents of D_v and B were significant at the 95% level, giving the equation

$$\frac{(x_{lt} - x_B)}{(6\Gamma_b/D_v)} = (1.104 \times 10^3) D_v^{0.64}/B^{0.52} \tag{25}$$

The experimental data for stripping length between 9 and 152 cm have been plotted in Fig. 8 against the values calculated from Eq. (25), with the respective ranges 9–14, 25–35, and 50–152 cm represented by different symbols. There is no difference among the different ranges,

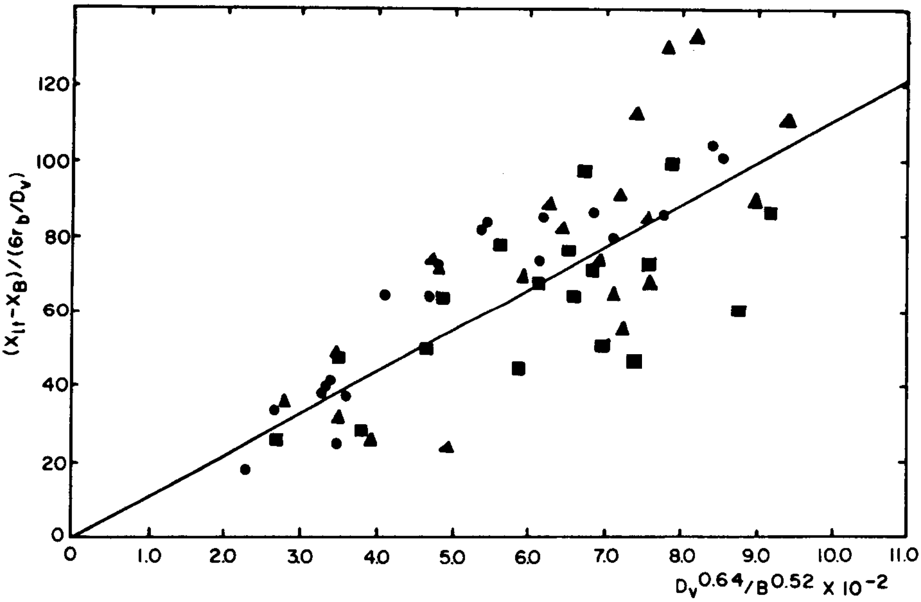

FIG. 8. $(x_{lt} - x_B)/(6\Gamma_b/D_v)$ as a function of D_v and B. Line: Equation (25), Points: experimental. Stripping column length: (●) 50–152 cm, (▲) 25–35 cm, and (■) 9–14 cm.

although the data do scatter. From the analysis it can be concluded that no effect of stripping column height is evident. Therefore, within the present experimental range the proposed mechanism involving negligible solute transfer within the foam is substantially correct for stripping column heights above approximately 10 cm. Under the present experimental conditions, 10 cm correspond approximately to the height of the mixing region below the feed. Consequently, below 10 cm stripping length may affect separation.

The fact that the concentration term could be correlated as a function of B and D_v and not F and G may be a reflection not so much of a physical relation but of the fact that F and G were not varied much because of limitations of the experimental system. Note that the present results cover relatively wide ranges of concentration (a factor of about 3), bottoms flow rate (a factor of 10), and column height (a factor of 30). Bubble size and other flow rates did not cover so wide a range. Thus care should be exercised in attempting to extrapolate the present results to other systems.

It should be pointed out that the present work is almost the first time that the height of a stripping column has been varied over a wide range in order to test the approach to operation as an infinite column. The one exception is the work of Haas and Johnson (*7*). They used the transfer unit approach, but their experimental results did not correspond to their theory. They were unable to predict the effect of flow rates and bubble size on HTU, and for stripping column heights over 28 cm their HTU values scattered badly. Most of their experimental results can be explained, at least qualitatively, with the presently proposed model of no solute transfer in the countercurrent region.

In accordance with this model, stripping length does not affect separation as long as the countercurrent region exists. Below a certain length, under given experimental conditions, the countercurrent region disappears, the two mixing regions merge, and foam height affects separation. It may be expected that length of the mixing regions depends on flow rates, bubble size, and efficiency of feed distribution.

Thus Haas and Johnson (*7*) report little variation in HTU values with stripping length of 10–28 cm at good operating conditions. The HTU values for 50, 60, and 85 cm stripping length scattered badly, and were usually larger than those for the shorter stripping length at otherwise similar conditions. These results, according to the presently proposed model, mean that for stripping length below 28 cm the two mixing regions merged (no countercurrent region) under the experimental conditions, and foam height affected separation. For 50, 60, and 85 cm stripping length the countercurrent region exists and separation is not improved, i.e., apparent HTU increases. Finally, as might be expected, they found that for short stripping length the type of liquid feed distributor affected separation.

Fanlo and Lemlich (*5*) present a curve of mass transfer coefficient as a function of calculated flows within the foam, but they evidently worked at a single column height only and assumed a linear effect of column height. In view of the present work, it is questionable whether their conclusions are valid for varying stripping column height.

CONCLUSIONS

Within the range of experimental variables studied in the present work, height of a stripping foam fractionation column of over 10 cm has practically no effect on separation. This experimental finding, indicating negligible solute transfer in most of the stripping length of the foam, cannot be explained satisfactorily by the transfer unit approach.

The proposed physical model which seems to explain the performance of stripping columns involves three regions: a mixing region involving primary the liquid pool at the bottom of the foam, a mixing region around the feed point, and a countercurrent region in between. Solute is transferred almost exclusively in the mixing regions. The length of the upper mixing region depends among others also on the method of feed distribution.

Stripping length does not affect separation as long as the countercurrent region exists. Below a certain stripping column height, under given experimental conditions, the countercurrent region disappears, the two mixing regions merge, and stripping column height affects separation.

SYMBOLS

a	surface area in foam, cm^2/cm^3 foam
A_c	column cross-sectional area, cm^2
D_v	average diameter of bubbles, cm
D_m	molecular diffusion coefficient
f	foam ratio, cc liquid/cc foam
F	feed flow rate, cc/min
G	gas flow rate, cc/min

HTU	height of a transfer unit, cm
L_d	flow rate of liquid traveling down within foam, cc/min
L_u	flow rate of liquid traveling up with foam, cc/min
NTU	number of transfer units
R'	universal gas constant, 8.314×10^4 erg/mmole °K
T	temperature, °K
V	flow rate of foam, cc/min
x	concentration, mole/cc
x_B	bottoms liquid concentration, mole/cc
x_F	feed concentration, mole/cc
y	concentration in collapsed foam, mole/cc
y_t	exit foam concentration, mole/cc
γ	surface tension, dyne/cm
Γ	surface excess, mole/cm^2

Acknowledgment

The financial assistance of the Ford Foundation is gratefully acknowledged.

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Received by editor May 5, 1971